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Development of Fluid Bed Gasification Control Parameters for a Range of Canadian Coals, Phase IV

Final Report

by

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ABSTRACT

The aim of the work reported herein was twofold, namely, (i) to perform fluid-bed coal gasification experiments and to compare results with analytical model results, and (ii) to study the details of the coal gas combustion field in a gas turbine type combustor.

In order to allow comparison with a model, the model had to be extended to apply to plug flow reactors, which the device under study is. When experiments were performed in an attempt to compare the results with analytical results, it was concluded that the present bench scale apparatus could not be expected to produce results which could be compared with analytical results due to inadequacies in the gasifier as it exists. These inadequacies include a large surface to volume ratio causing excessive heat losses; a coal auger whose feeding rate depends on the coal depth in the tank supplying it, thereby defying predictability of its operation; an ash removal system which did not behave as planned; and instrumentation which failed to provide reliable steam flow measurements. Recommendations were made for a better design which could overcome these problems.

Partial completion of the extensive study of combustion of coal gas in a gas turbine type combustor was reported. Results are presented from temperature measurements using an aspirated thermocouple probe, and velocity vector studies using both a two-hole and a five-hole probe.

Three components of velocity are obtained. The first few sections of the combustor showed recirculation zones and related sharp gradients in flow parameters. Thereafter, the axial velocity became quite stable, although still varying, and circumferential and radial velocity components persisted to the combustor exit, for the conditions studied to date.

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The work described herein was coordinated through Asecor Ltd., with experimental testing being carried on at Carleton University where the gasifier and combustor are located. The work was supported, in part, by NSERC Grant awarded to Prof. E.G. Plett, Department of Mechanical and Aeronautical Engineering, Carleton University.

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1.0 INTRODUCTION

1.1 Status at the Start of this Work

The work reported herein is based on the fourth year of a study aimed at the development and utilization of a bench scale fluidized bed coal gasifier and gas burner system to be used to study the gasification behaviour of various Canadian coals. The initial development of the bench scale system is described in some detail in Reference 1, with some modifications described in Reference 2. Some preliminary results with the system were reported in the first two annual reports, but only limited data were obtained. Some further results were obtained during the third year[3], but work continued to obtain even more results in the fourth year work, which is the subject of this report.

The general dimensions and configuration of the fluidized bed gasifier as developed during the first year of this research program are shown in Fig. 1. During the second year, changes of some importance included a new coal feeder, ash pit and fly ash return as shown schematically in Fig. 2. In addition, some improvement in the method for ignition was developed during the second year.

In parallel with the gasifier development, a combustor was developed to permit studies of the combustion behaviour of low to medium heating value mixtures of gases such as obtained from a fluidized bed gasifier. Fig. 3 shows an assembly drawing of the combustor designed for that purpose. The design was made during the second year as described in the second year report[2].

A theoretical model was developed during the third year in conjunction with the objective of optimizing performance of the gasifier. Considerable results were reported based on that model. Further experimental work was undertaken during this fourth year for the purpose of verifying the theory. This is reported on in this report.

1.2 Objectives of the Fourth Year Study

There were two overall objectives to be pursued during this fourth year of the study. Building on the work of the previous three years, the objectives were:

a) to perform fluid bed coal gasification experiments and to compare results with analytical model results;

b) to study the details of the coal gas combustion field in the gas turbine type combustor, to relate the combustion characteristics to the flame structure.

2.0 FLUID BED COAL GASIFICATION

2.1 Plug Flow Reactor Model

In a fluidized bed facility, if the active bed height is more than four bed diameters, then the bed behaves as a plug-flow. It is unlikely that any commercial sized unit would fall into this regime, due to the large diameters contemplated, but small scale reactors often do. A small fluid-bed gasifier may be treated as a back-mixed or plug-flow reactor, depending on the solids inventory of the bed.

In the course of this work, a plug-flow reactor (PFR) model with the same geometry as a fully back-mixed fluidized bed reactor (BMR) were compared. This is an academic comparison, in a way, since the large unit does not satisfy the bed height to diameter ratio of larger than four, as required for plug flow. The comparison does yield some interesting insight into the differences between the two regimes for the same size reactor with the same coal and feed rates.

Table III compares the results of three runs for each model with identical independent process variables. Some characteristic differences are:

- the carbon conversion ratio and thermal efficiencies according to the PFR model are much higher than those according to the BMR model. This agrees well with the general experience on the performance of plug-flow reactors.
- the PFR model predicts consistently higher CH_4 and lower H_2 contents of the syngas and due to the large difference between the heating values of these two gases, the heating values of the syngas are considerably higher according to the PFR model.

- because of the higher efficiencies predicted by the PFR model, the power output of the unit, the syngas temperature and velocities in the reactor are higher too. The residence time of coal in the bed is also higher due to the lower bed ash content.

In general, a plug-flow reactor will have a better performance compared to a well-mixed reactor. Since bench scale and even pilot plant gasifiers operate usually as plug-flow reactors while commercial gasifiers operate as back-mixed reactors, a simple extrapolation of test results of the former will give an erroneous prediction of the performance of the latter. This is avoided in an analytical study where the phenomena occurring in the gasifier are properly modelled.

Figure 4 shows the syngas composition and progress of reactions in the fluid bed according to both models. It is noticeable that in a plug flow reactor the base carbon conversion ratio is very high in the region of the bed below the coal inlet(s), falling slightly below its back-mixed value close to the top of the bed. Notice also that the rate of H_2O reduction and H_2,CO and CH_4 increase is high in the zone of coal inlet where the rapid gasification reactions take place. Finally, the position of the coal feed lines relative to the gas distributor may considerably affect the performance of a small reactor but in a much lesser degree the performance of a big gasifier.

2.2 Comparison of Experimental and Analytical Results

In the three reports on the first three years' work on this project[1-3], preliminary results were shown for the operation of the small, bench scale fluid bed gasifier located at Carleton University. During the past year, further effort was directed toward obtaining more detailed data to be used in comparing the theory with experimental results. In order to make a meaningful comparison, it is important that steady state can be achieved in the gasifier for a reasonable period of time, and that the conditions could be well documented.

Based on this latest, most critical testing period as well as earlier experience[3], it has been concluded that the existing unit is not

suitable for steady state operation as a fluidized bed gasifier for sufficiently long periods of time to make the necessary tests to compare with analytical results. Several reasons for this inadequacy are discussed in the following subsections.

2.2.1 Effect of size of unit

The diameter of this gasifier is quite small (7 cm I.D.) leading to a large surface/volume ratio (~57 m^{-1}). Thus, despite the use of 20 cm thick insulation around the outside, the heat losses are excessive, making it impossible to maintain high enough temperatures. Typically, the gas temperature can be 700-1000°C in the bed and drop to 50-150°C above the bed in the free board. In addition, due to the relatively large amount of metal surrounding the inner area, a very prolonged heating period is required to bring the walls to a sufficiently high temperature to commence attempts to bring the unit into a normal operation.

For example, after from 1 to 2 hours of preheating with CO, coal and 0_2 (the 0_2 was preheated to ~150°C), the bed temperature reaches between 800 and 1000°C while the gas exit temperature is only 100-150°C. Shutting off the CO supply at this point results in a decrease in bed temperature to 500-550°C within about 10 minutes. Attempts to raise this temperature without re-introducing CO or some other fuel gas, but only varying coal (Byron creek coal) and oxygen flow rates failed to raise the bed temperature above 600°C. Substituting air for 0_2 as the oxidizing gas would only serve to lower the temperature further. Then, when steam at about 180°C is introduced slowly so that gasification reactions could begin, the bed temperature is further lowered to such a low point that the necessary reactions cannot be sustained. Due to the endothermicity of the gasification reactions, gasification was only possible when CO was burned to raise the bed temperature to above 700°C. This was earlier found to be necessary with Estevan lignite as well[3].

The use of CO combustion as an additional source of heat within the bed introduces other problems. The CO was introduced through the igniter above the distributor plate, from which it flowed upwards in a plug flow

but with most of it burning well above the distributor plate and some even in the free board. This generated a very non-uniform temperature profile in the bed, with temperatures varying by between 100 and 150°C. The free board conditions were also now far from that of a conventional fluid bed gasifier. Therefore, no practical use could be made of either result. In addition, the combustion of the CO required a larger amount of 0_2 than is otherwise required for fluidized bed gasifiers. The overall result is that this non-uniform, unusual situation cannot be simulated by an analytical model intended to simulate a fluidized bed gasifier, and therefore offers very little value in terms of comparisons between experimental results and the results of analytical models which have been developed.

2.2.2 Coal feeding problems

The present coal feeder is a screw conveyor using only mechanical feeding principles. It provides two basic problems:

(a) Gases flowing through the main gasifier unit can enter the coal hopper via the feed tube. Consequently, when steam is introduced, it can enter the coal feeder and hopper where it condenses due to the lower temperature in these zones. During a given run, this does not cause as great a problem as between runs, although the steam condensing in the hopper is far from the bed where it is assumed to be causing gasification reactions. Between runs, however, any coal left in the hopper must be completely removed and the entire system carefully dried (a tedious task), or the coal becomes a solid mass of unknown moisture content and hardness like the original unground coal.

(b) With the simple screw feeder, the coal feed rate is strongly dependent on the amount of coal in the hopper, on the pressure in the gasifier and on the coal moisture content, which varies as steam condenses, as mentioned above. The overall result is that the feed rate, with constant screw rotation rate, varies so widely that mass balances cannot be made or the residence times in the bed computed with any degree of certainty.

2.2.3 Ash removal limitations

With the present arrangement, continuous ash removal at a rate to control the bed size is not possible. Ash can be removed in a batch operation by a downcomer, but this bed was designed to operate as an agglomerating ash gasifier. Under current operating conditions, the ash does not agglomerate, so the removal doesn't occur. Cleaning of the current ash hopper is also a very tedious and time-consuming operation which requires that the unit first be shut down and allowed to cool.

2.2.4 Instrumentation and data acquisition limitations

There are several problems with the instrumentation, some of which stem from problems with the operation, some contribute additional problems. In summary, these are the concerns:

(a) The coal feed rate cannot be accurately known due to the problems with the feeder.

(b) The steam flow rate measurement is unreliable even though the components were acquired, as advertized, specifically for this purpose.

(c) Sample analysis: the samples of dried syngas can be analyzed quite reliably in the chromatograph. Condensate amounts need to be collected over a measured time period but since the overall operation does not attain a proper steady state, this does not permit reliable condensate measurements, so that analysis of the condensate does not give a proper correspondence to the dried syngas measurement.

(d) The sampling probe broke down so that sampling of solids (which was never demonstrated) or gases from various parts of the bed and freeboard was never possible. In principle, this could be repaired if this was the only impairment to good data, but with the other more intrinsic problems not solved, this was not undertaken.

(e) No provision is in place to measure the bed height directly.

2.3 Recommendations

Due to the substantial problems which are inherent in the current system, it is recommended that a new design be developed based on experience gained with the current unit and results in related work by others. It is to be expected that a slightly larger unit is to be desired with several features with slightly different design incorporated into it. Briefly, the new design should incorporate the following features:

(a) Size: a square cross-section of about 40 cm per side and a height of about 2 m would be suitable.

(b) Insulation: an inner lining of refractory with outside layers of fiberfax and fiberglass in that order.

(c) Coal Feed: the coal feed system should include: a cylindrical hopper with a conical bottom leading to a rotary air lock valve with an air/0_2 blower to feed the coal from an inlet protruding through the bed bottom. The blower delivery pressure should be adjusted to 3-5 psig above the pressure at the bottom of the fluid bed. Conveying air/0_2 to coal ratio should be between 1:5 and 1:10.

(d) Ash Removal: a system capable of continuous ash removal from the bed at a controllable rate should be included. It could consist of a water cooled screw conveyor, rotary air lock valve and ash hopper.

(e) Distributor: a gas distributor with bubble cap and side holes should be used.

(f) Steam Supply: a steam superheater capable of delivering 300°C steam should be added to the boiler outlet.

(g) Air/0_2 Preheater: a properly sized preheater for the fluidizing gas capable of heating to at least 300°C is desirable.

(h) Igniter Combustor: a more powerful igniter - combustor for preheating the unit and igniting the bed is desired. It could operate on natural gas, to avoid the need for bottled fuel gas, and capable of being isolated from the gasifier when the gasifier attains self-sustained operation. The combustor flame would be outside the gasifier so that the high temperature

exhaust gases serve as the heat transfer fluid. The entrance should be very near the distributor plate, or in part, even below the distributor to raise its temperature as well.

(i) Cyclone Separator: the present cyclone seems adequate for this bed, except that positive means must be used to bring the fly ash back into the system. An air lock valve at the base of the cyclone together with a small blower could be used to re-introduce the fly ash to the bed via the coal feed stream. Provision for sampling the fly ash should be included.

(j) Syngas Cooler: the current system has this, but a slightly refined design would be desirable.

(k) Thermocouples: at least ten ports along the length of the gasifier would be more reliable than trying to use a traversing probe to perform too many functions.

(l) Sampling Probe: a vertically traversing gas and solids sampling probe inserted from the top of the bed, much like our current design is recommended.

(m) Static Pressure Ports: at least three but possibly more pressure ports with perhaps a rotary valve leading to a readout device would be needed to measure pressure changes through the bed.

(n) Bed Height Probes: heat transfer probes in the vicinity of the bed top could be used to monitor bed height at least approximately, since the heat transfer rate within the solids region is much higher than in the gaseous regime above the bed.

These are some of the areas where we feel improvements could be made. Figure 5 depicts a possible layout of the new design. Since design and building of such a new unit was not included within the scope of the reported work, no work beyond the above recommendations was undertaken relating to it.

3.0 COAL GAS COMBUSTOR STUDIES

The purpose of these studies was to develop a more complete understanding of the various aspects of the problem of burning low to medium heating value gases in a gas turbine type combustor. The relationship to the gasifier work is in the idea that coal could be gasified, have the gases cleaned to meet environmental constraints and then burned in a gas turbine for power generation. Consequently, gases as produced from the gasifier would be burned. An oxygen blown gasifier could be expected to produce medium heating value gases, without upgrading, whereas an air blown gasifier would produce low heating value gases. Hence, the range of interest.

In this report, some experimental results relating to thermal and fluid mechanics studies inside the test combustor are presented. Gas analysis studies with thermal analysis are being used to obtain an independent check of temperature in the flow. Results of these studies will be reported later. The main emphasis of the work reported herein relates to the effects of certain parameters on the performance of the combustor and the effect of these parameters on the combustion space and the various zones encountered in industrial combustors. A final goal of this work is to provide information to allow better formulation of combustion models for combustor analysis, as applied to low to medium heating value fuels.

3.1 Test Rig and Flow Measurement

Figure 6 shows a schematic diagram of the arrangement used for the experiments being reported. It consists of a combustor, gaseous fuel supply arrangement and air supply with provisions for metering each flow. The fuel was obtained by blending CO, H_2 and CO_2 from gas bottles. The heating value was varied by changing the percentage of CO_2. The blend of CO, H_2 and CO_2 used represents the proportion to be expected for the gas produced from a fluidized bed oxygen blown coal gasifier. The maximum heating value which could be obtained using mixtures of CO and H_2 is about 11 MJ/m³ (300 Btu/scf). In order to obtain higher values, methane

(CH_4), (36 MJ/m³ or 1000 Btu/scf) must be included in the blend. The amount of fuel supplied to the combustor is measured by three separate rotameters, one for each gas.

A centrifugal blower of air capacity up to 1857 m³/hr at 3100 rpm is used to supply the air. To measure the air delivered to the combustor and also to allow control of the amount of air supplied to each section of the combustor, flow metering devices are used here as well. The total air supply is measured using a venturi meter connected to an inclined manometer. The air is then divided into two parts, one is supplied directly to the combustor dilution zone, the other part is connected to six flow meters, one for the premixed air, one for the primary air and four meters for the secondary zone of the combustor. Fig. 3 shows a cross-sectional view of the combustor whose design was described in Reference 4.

3.2 Instrumentation

3.2.1 Thermocouple

An aspirated pyrometer probe is used to measure the temperature at each section of the combustor. Figure 7 shows a cross-section of the suction pyrometer mounted thermocouple. The thermocouple is made of fine Pt-13% Rh wire of 0.2 mm diameter inserted in a 2-hole insulator. Aspiration of the gas is achieved by using a suction pump, and adjusting the flow at each section to match the average velocity in that section.

3.2.2 Two-hole probe

Figure 8 shows the water cooled two-hole probe designed for this work. It precedes the use of the 5-hole probe for velocity measurements. This probe determines the direction of the flow and identifies the recirculation zone.

3.2.3 Five-hole probe

A water cooled 5-hole probe is used to obtain the velocity vector at selected sections of the combustor. The probe is type DA-375 manufactured by United Sensors. Using inclined manometers, readings for the values

of P_1-P_{atm}, P_1-P_2, and P_4-P_5 (numbers 1-5 refer to the five holes at the tip of the probe) were obtained. The probe is rotated about its axis until P_2-P_3 vanishes and that gives the yaw angle between the velocity vector and the probe axis. Using the calibration information together with the manometer readings allowed the velocity vector to be obtained.

3.3 Results

Figure 9 shows a sample result of the tracing by the 2-hole probe. It indicates that the recirculation zone extends only down to the third section. Figure 10 is a sample result of the temperature distribution across the combustor at intervals along it. Figure 11 shows the temperature distribution at the exit section at four different running conditions.

Figure 12 gives a sample result of the axial velocity component distribution throughout the combustor. Figure 13 shows the axial velocity component at the exit section at four different testing conditions.

Figure 14 gives a sample result of the radial velocity component distribution while Fig. 15 gives the radial component at the exit section for four different testing conditions. Figures 16 and 17 give similar results concerning the tangential component of velocity.

3.4 Discussion

All the results given in this report are obtained from experiments carried out using one swirler, hence the effect of swirl, if any, can not be identified at this time. However, the results show a large variation in changing from one operating condition to another. Temperature measurement showed that it is possible to obtain the desired shape of distribution through the choice of the operating conditions. It also indicated positions of hot spots in the flame tube where better cooling is needed at this section or improved air admittance is necessary for the different velocity components. With respect to velocity components, it is observed that the size of the recirculation zone which determines the size of the combustion zone can be varied by varying operating conditions. This is of particular interest in the present case, in the interests of expanding

the combustion zone for these type of fuels which are known for their tendency to have a short zone due to their heat content and chemical content. Also, the velocity profile of the axial component at the exit section could be adjusted by the choice of the operating conditions and/or the design of the exit section itself, to meet the need of a turbine stage downstream of the combustor. Measurement of velocity using a 5-hole probe gave almost no problems until the probe was inserted into the high temperature in the first three sections of the combustor. It seems that the difficulty arose due to condensation of water vapour, resulting from the combustion of H_2, inside the cooled probe passage, which resulted in distortion of manometer readings. However, cooling the probe with steam and purging it after every reading with air is a solution to this problem.

It is a known fact that static pressure at any section will rise due to insertion of an object in this section which reduces the area of the section. This could be corrected for by using the variation of area ratio calibration with the change in static pressure. The total pressure measured at the centre of the probe tip could also be in error as a result of either the pitch angle or the yaw angle. The probe is calibrated for these effects.

The results show clearly that properties inside the combustor are not symmetric in any way, nor are they predictable from one section to another. This could be seen in the temperature distribution, Fig. 10. The temperature at any point in the upper half of the combustor is generally higher than the corresponding point in the lower half. This may be explained in terms of buoyancy of the hot gases associated with the flame zone. As for velocity components, the axial component gave somewhat consistent values and trends from one test to another. The axial velocity variation across a given section was very little, except in the first three sections near the flow inlet and except within the first 5 mm from the wall. For the other components, there was more consistency from one test to another and also some steep gradients at some sections. Also, it was of interest to note that the circumferential and radial components were observed throughout the combustor, to the exit section.

3.5 Conclusions

In conclusion, it must be mentioned that a complete model which will define the combustion process inside a gas turbine combustor must include the three-dimensional change of almost all the defining properties. Also, cooling of the combustor liner must be included in these models or have a model of its own due to its effect on the velocity, in particular near the combustor walls as seen in some results as a sudden rise in axial velocity component when it is measured near a point where air cooling was admitted. The technique which is used to admit air for combustion or dilution must also be carefully modelled due to its profound effect on the flow field. The position of the holes to admit this air must also be optimized for combustor performance in particular if a desired shape of axial velocity and temperature profiles at the exit section is to be achieved.

At this stage, although the results are incomplete, and a final recommendation or conclusion for the optimum operating conditions is not possible, it is possible to note that variation of the parameters chosen produced real and clear change. In the work that is continuing, variation of the parameters of interest will be completed, and their effect will be fully documented and comparisons made. The feedback of these conclusions in terms of better models for combustion and combustion equipment and their design for this type of fuel will be outlined then.

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TABLE I

PROPERTIES OF COALS AS CONSIDERED IN THE CASE STUDIES

PROPERTIES	UNITS	COAL #1	COAL #2	COAL #3
		BITUMIN. HI-VOL.A	SUBBITUM. C	LIGNITE
Fixed Carbon	-	0.4700	0.3840	0.2990
Volatile Matter	-	0.3209	0.2520	0.2470
Ash	-	0.1830	0.1670	0.0990
Moisture	-	0.0261	0.1970	0.3550
Carbon	-	0.6470	0.4719	0.3931
Hydrogen	-	0.0419	0.0312	0.0262
Sulfur	-	0.0680	0.0019	0.0027
Nitrogen	-	0.0071	0.0057	0.0071
Oxygen	-	0.0269	0.1253	0.1169
Higher Heating Value	kJ/kg	26320.	18200	15310
True Density	kg/m^3	1200	1200	1200
Mean Particle Size	μm	500	500	500
Particle Size, dp_{10}	μm	400	400	400
Initial Sintering Temp.	K	>1600	>1600	>1600
Coal Pretreatment Temp.	K	<500	<500	<500
Voidage at Min. Fluidization	-	0.55	0.55	0.55
Reactivity Factor, f_R	-	1.5	1.5	1.5
Reactivity Factor, f_o	-	.9228	1.187	10.0

TABLE II

OPTIMUM PERFORMANCE OF A GIVEN GASIFIER FOR DIFFERENT COALS

P_H = 400 MW; OPERATION ACCORDING TO CASE A.

#	GROUP	P A R A M E T E R	SYMBOL	COMPUTER SYMBOL	UNIT	COAL A	COAL B	COAL C
1	GEOMETRICAL-PHYSICAL PARAMETERS	Diameter of gasifier	D_g	DG	m	6.		
2		Internal height of gasifier	L_T	BLT	m	40.		
3		Number of holes of distributor plate	N_D	BND	-	120000.		
4		Diameter of holes	D_h	DH	m	0.008		
5		Thickness of distributor plate	l_h	DISTHI	m	0.05		
6		Thickness of insulation of the gasifier	Δt_{INS}	THICKN	m	0.10		
7		Thermal conductivity of insulation	k_{INS}	BKINSU	kW/m.K	0.0001		
8		Outside convective heat-transfer coefficient	h_o	HOUT	$kW/m^2.K$	0.20		
9		Overall heat-transfer coefficient	U	U	$kW/m^2.K$	$0.995.10^{-3}$		
10		Cross-sectional area of gas distributor	A_t	AT	m^2	28.274		
11		Internal heat-transferring area of FBG	A	A	m^2	728.256		
12		Overall collecting efficiency of cyclone(s)	η_{CN}	HCN	-	0.90		
13		Max. allowed particles in cleaned syngas	F_{PS}	FPS	kg/m^3	10^-		
14		Ambient temperature	T_e	TE	K	300.0		
15	COAL PROPERTIES	Higher heating value of coal	CV_{COAL}	CV1	kJ/kg	26320.	18200.	15310.
16		True density of raw coal	ρ_1	RO1	kg/m^3	1200.	1200.	1200.
17		Coal composition: H_2O	-	WATER	%	2.61	19.70	35.50
18		" " : H_2	-	HYGROG	%	4.19	3.12	2.62
19		" " : C	-	CARBON	%	64.70	47.19	39.31

continued....

20	COAL PROPERTIES	Coal composition: N_2	-	BNITRO	%	0.71	0.57	0.71
21		" " : S	-	SULFUR	%	6.80	0.19	0.27
22		" " : O_2	-	OXYGEN	%	2.69	12.53	11.69
23		" " : Ash	-	ASH	%	18.30	16.70	9.90
24		Base carbon in coal	M_{cb}	BMCB	kmole/kg	.0376	.0307	.0239
25		Volatile carbon in coal	M_{cv}	BMCV	"	.0163	.0086	.0088
26		Volatile matter in coal	-	-	%	32.09	25.20	24.70
27		Fixed carbon in coal	-	FIXCAR	%	47.00	38.40	29.90
28		Mean diameter of coal particles	$\bar{d}_p$	DP	m	.0005	0.0005	0.0005
29		Specially defined particle diameter	d_{p10}	DP10	m	.0004	0.0004	0.0004
30		Initial coal sintering temperature	T_s	TS	K	2000.	2000.	2000.
31		Coal pretreatment temperature	T_p	TP	K	500.	500.	500.
32		Coal reactivity: f_o	f_o	F0	-	0.923	1.187	10.0
33		Coal reactivity: f_L	f_L	FL	-	44.84	55.83	587.6
34		Coal reactivity: f_R	f_R	FR	-	1.50	1.50	1.50
35	INDEPENDENT PROCESS VARIABLES	Mass of bed inventory	W	Z	kg	295700	454300	417200
36		Coal feed rate	m_c	BMC	kg/sec	12.119	19.522	24.044
37		Oxygen feed rate	M'_{O_2}	BMP5	kmole/kg_c	.01805	.01557	.01313
38		Nitrogen feed rate	M'_{N_2}	BMP7	"	.06789	.05855	.04191
39		Steam feed rate	M'_{H_2O}	BMP8	"	.08125	.02547	.01460
40		Pressure in gas mixing vessel	p_m	PM	kPa	478.	590.	568.5
41		Temperature of oxygen/air	T_a	TA	K	850.	850.	850.
42		Temperature of steam	T_w	TW	K	850.	431.	850.

continued....

43	PERFORMANCE PARAMETERS	Syngas composition : H_2	y_{H_2}	PSI4	%	19.06	15.25	16.82
44		" " : N_2	y_{N_2}	PSI7	%	31.67	40.71	35.74
45		" " : H_2O	y_{H_2O}	PSI8	%	23.46	16.76	19.60
46		" " : H_2S	y_{H_2S}	PSI14	%	0.74	0.03	0.054
47		" " : CH_4	y_{CH_4}	PSI11	%	2.45	2.00	1.88
48		" " : CO_2	y_{CO_2}	PSI12	%	9.94	10.23	12.04
49		" " : CO	y_{CO}	PSI13	%	12.67	15.02	13.86
50		Heating value of syngas	CV_H	CVH	kJ/m^3	3936.	4260.	4430.
51		" " "	CV_{H0}	CVH0	kJ/SCM	4589.	4250.	4250.
52		" " "	CV_L	CVL	kJ/m^3	3566.	3913.	4045.
53		" " "	CV_{L0}	CVL0	kJ/SCM	4158.	3904.	3881.
54		Sensible enthalpy of syngas	HC	HC	kJ/m^3	1445.	1559.	1582.
55		Power output of gasifier	P_H	PH	kW	400000.	400000.	400000.
56		" " "	P_L	PL	kW	372512.	376135.	374400.
57		" " "	P_{HCOLD}	PHCOLD	kW	292581.	292790.	294734.
58		" " "	P_{LCOLD}	PLCOLD	kW	265108.	268949.	269119.
59		Power concentration of gasifier	PC_H	PCH	kW/m^3	353.66	353.66	353.66
60		" " "	PC_L	PCL	kW/m^3	329.37	332.57	331.04
61		Carbon conversion ratio	η_C	HCCR	-	1.00	1.00	1.00
62		Auxiliary power consumption	E_{AUX}	EAUX	kJ/kg-coal	89.86	135.68	252.09
63		Thermal efficiency of the gasifier	η_H	HTHH	%	98.18	97.86	94.92
64		" " "	η_L	HTHL	%	91.43	92.03	88.84
65		" " "	η_{HCOLD}	HHCOLD	%	71.81	71.63	69.93
66		" " "	η_{LCOLD}	HLCOLD	%	65.07	65.80	63.86
67		Pure oxygen feed rate	-	OXYPRO	$kmole/kg_C$	0.0	0.0	0.0
68		External heat input	Q_{INPUT}	QINPUT	kJ/kg-coal	0.0	0.0	0.0

Continued....

69		Temperature of gasifying media	T_m	TM	K	850.	728.	850.
70		Temperature of fluid-bed and syngas	T	T	K	1211.	1221.5	1155.
71		Pressure drop across the distributor	Δp_d	DPD	kPa	22.28	15.76	14.47
72		" " " the fluid-bed	Δp_b	DPB	kPa	102.59	157.62	144.75
73		Pressure of syngas	p_g	PG	kPa	353.13	416.62	409.28
74		Gas velocity below gas distributor	$(u_o)_o$	U00	m/sec	1.059	0.705	0.736
75		" " above " "	u_o	U0	m/sec	1.111	0.725	0.755
76		" " in free board†	u_2	U2	m/sec	2.629	2.431	2.353
77	DEPENDENT PROCESS VARIABLES	Min. fluidization velocity	u_{mf}	UMF	m/sec	0.226	0.226	0.210
78		Min. velocity for bubbling fluidization	u_{pf}	UPF	m/sec	0.452	0.452	0.420
79		Terminal velocity of particles	u_T	UT	m/sec	2.629	2.450	2.465
80		Density of gasifying medium	ρ_{FM}	ROG	kg/m^3	1.520	2.471	2.090
81		" of syngas	ρ_{SG}	RO2	"	0.773	0.974	0.996
82		" of the bed (true)	ρ_S	RO16	"	2700.	2700.	2700.
83		Active bed height	L_f	BLF	m	10.96	15.46	14.54
84		Residence time of coal in bed*	τ	TIME	sec	133331	139348	175267
85		Incipient bed voidage	ε_{mf}	EMF	-	0.55	0.55	0.55

† free board area is the same as the bed area

* See Appendix I for comparison of these steady state, optimized results, with available results in the literature.

TABLE III

COMPARISON OF THE BACKMIXED AND PLUG-FLOW REACTOR MODELS

(Coal properties as in Table I, FBG geometry as in Table II, bituminous coal)

#	GROUP	SYMBOL	UNIT	RUN # 1		RUN # 2		RUN # 3	
				BACKMIXED	PLUG FLOW	BACKMIXED	PLUG FLOW	BACKMIXED	PLUG FLOW
1	INDEPENDENT PROCESS VARIABLES	W	kg	291900.0	291900.0	291900.0	291900.0	291900.0	291900.0
2		m_c	kg/sec	16.738	16.738	16.7380	16.7380	16.7380	16.7380
3		M'_{O_2}	$kmole/kg_c$	0.010	0.0100	0.0120	0.0120	0.0120	0.0120
4		M'_{N_2}	"	0.0376	0.0376	0.04512	0.04512	0.04512	0.04512
5		M'_{H_2O}	"	0.09441	0.09441	0.09441	0.09441	0.09441	0.09441
6		p_m	kPa	622.04	622.04	622.04	622.04	622.04	622.04
7		T_a	K	450.0	450.0	450.0	450.0	850.0	850.0
8		T_w	K	450.0	450.0	450.0	450.0	850.0	850.0
9	PERFORMANCE PARAMETERS	y_{H_2}	%	13.09	6.74	12.62	7.38	13.09	6.74
10		y_{N_2}	%	22.39	23.22	25.34	26.11	22.40	23.23
11		y_{H_2O}	%	46.44	44.03	43.87	41.63	46.44	44.03
12		y_{H_2S}	%	0.94	0.98	0.89	0.92	0.94	0.98
13		y_{CH_4}	%	4.32	9.96	3.93	8.64	4.32	9.96
14		y_{CO_2}	%	10.29	13.03	10.67	12.98	10.29	13.03
15		y_{CO}	%	2.52	2.03	2.68	2.335	2.52	2.03
16		CV_H	kJ/m^3	4944.9	6491.6	4668.3	5959.6	4944.9	6491.6
17		CV_{HO}	kJ/SCM	3394.2	4652.0	3216.8	4282.0	3394.2	4652.0
18		CV_L	kJ/m^3	4376.0	5822.9	4133.9	5342.5	4376.0	5822.9

continued....

#	GROUP	SYMBOL	UNIT	RUN # 1		RUN # 2		RUN # 3	
				BACKMIXED	PLUG FLOW	BACKMIXED	PLUG FLOW	BACKMIXED	PLUG FLOW
19	PERFORMANCE PARAMETERS	CV_{LO}	kJ/SCM	3005.7	4172.7	2848.5	3838.6	2942.9	3756.5
20		HC	kJ/m^3	2754.2	2754.7	2666.1	2660.1	2547.9	2410.2
21		P_H	kW	365551.4	441956.7	370316.8	440500.5	385930.9	455622.8
22		P_L	kW	338539.4	409990.3	343333.8	408965.3	357965.9	421114.8
23		P_{HCOLD}	kW	234784.3	310285.1	235705.3	304557.8	246292.4	319010.4
24		P_{LCOLD}	kW	207772.3	278318.8	208722.3	273022.6	218327.4	284502.4
25		PC_H	kW/m^3	323.2	390.77	327.43	389.49	341.24	402.86
26		PC_L	"	299.33	362.51	303.57	361.60	316.51	372.35
27		η_c	-	0.4950	0.7144	0.5318	0.7306	0.5465	0.7080
28		E_{AUX}	kJ/kg-coal	99.65	99.65	119.58	119.58	119.58	119.58
29		η_H	%	68.85	83.25	69.62	82.81	67.63	79.85
30		η_L	%	63.77	77.23	64.55	76.88	62.73	73.80
31		η_{HCOLD}	%	44.23	58.45	44.31	57.26	43.16	55.90
32		η_{LCOLD}	%	39.13	52.43	39.24	51.33	38.26	49.85
33		Q_{INPUT}	kJ/kg-coal	0.0	0.0	0.0	0.0	0.0	0.0
34	DEPENDENT PROCESS VARIABLES	T_m	K	450.0	450.0	450.0	450.0	850.0	850.0
35		T	K	1028.1	1073.3	1028.3	1072.3	1076.8	1090.8
36		ΔP_d	kPa	11.45	11.45	13.31	13.31	25.14	25.14
37		ΔP_b	kPa	101.28	101.28	101.28	101.28	101.28	101.28
38		P_g	kPa	509.31	509.31	507.45	507.45	495.62	495.62
39		$(u_o)_o$	m/sec	0.515	0.515				
40		u_o	"	0.515	0.515	0.551	0.551	1.062	1.062
41		u_2	"	1.679	1.690	1.786	1.807	1.938	2.005
42		u_{mf}	"	0.211	0.224	0.209	0.222	0.1555	0.164

continued....

#	GROUP	SYMBOL	UNIT	RUN # 1		RUN # 2		RUN # 3	
				BACKMIXED	PLUG FLOW	BACKMIXED	PLUG FLOW	BACKMIXED	PLUG FLOW
43	DEPENDENT PROCESS VARIABLES	u_{pf}	m/sec	0.422	0.449	0.419	0.445	0.311	0.329
44		u_T	"	1.745	1.844	1.759	1.841	1.766	1.838
45		ρ_{FM}	kg/m^3	3.530	3.530	3.593	3.593	1.865	1.865
46		ρ_{SG}	"	1.259	1.300	1.283	1.310	1.185	1.176
47		ρ_S	"	1755.4	1899.0	1772.4	1915.5	1779.8	1892.9
48		L_f	m	14.39	13.23	14.39	13.24	17.14	15.96
49		τ	sec	34217.0	47426.	35895.	48813.	36617.	46903.
50		ϵ_{mf}	-	0.55	0.55	0.55	0.55	0.55	0.55

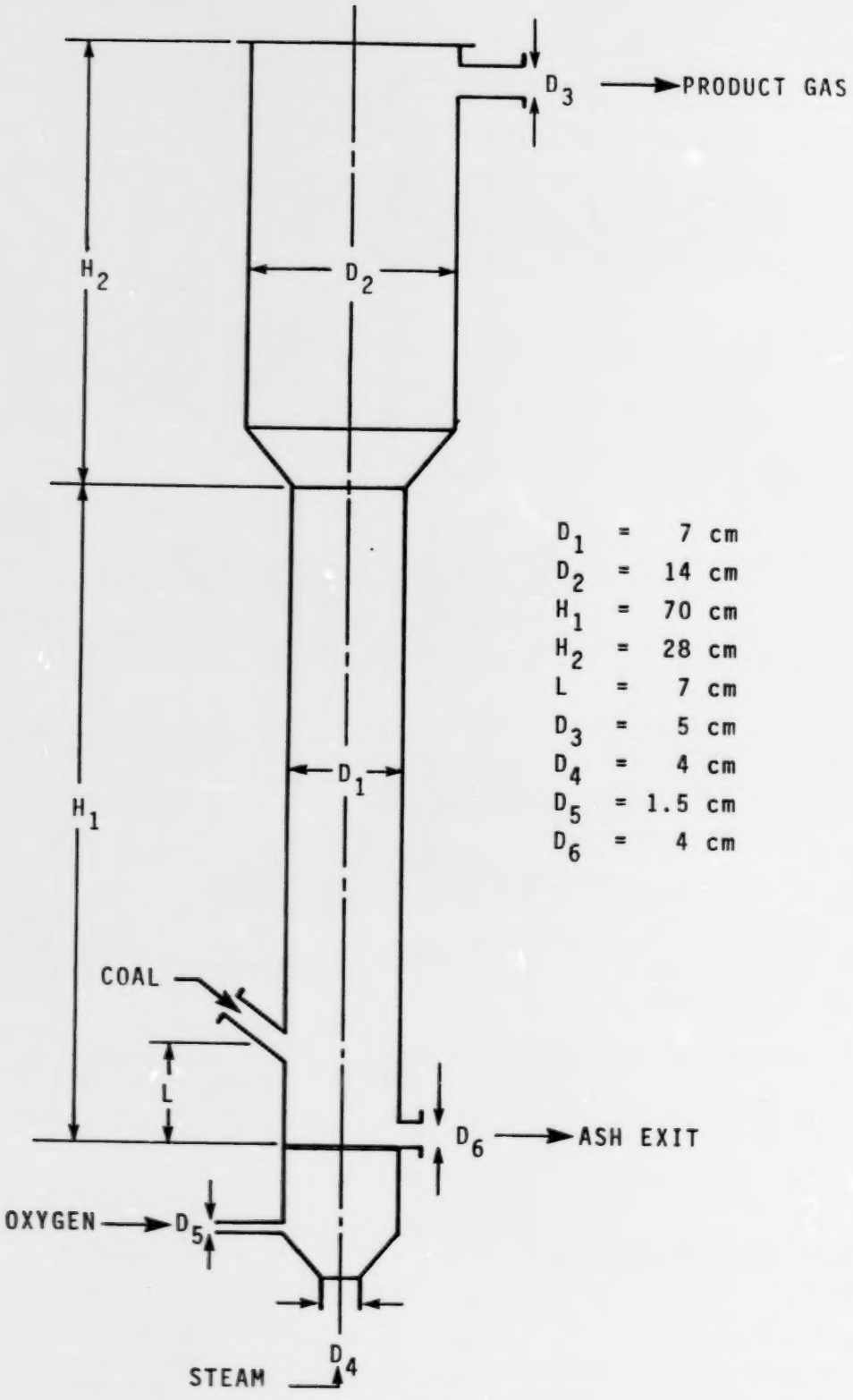


Figure 1: General Dimensions of the Fluidized Bed Gasifier as First Built

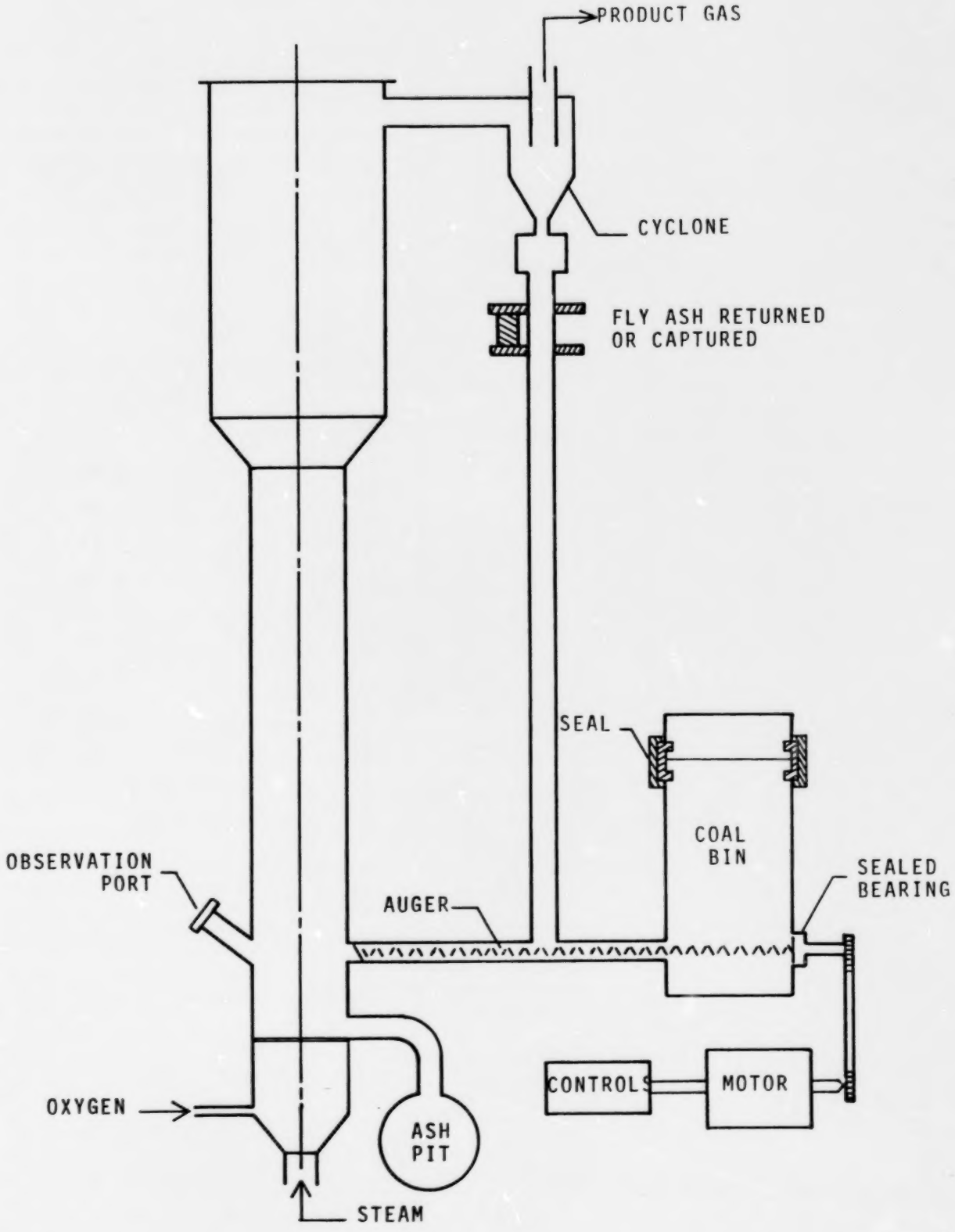


Figure 2 Gasifier showing coal feeder and ash pit which were added during the second contract period

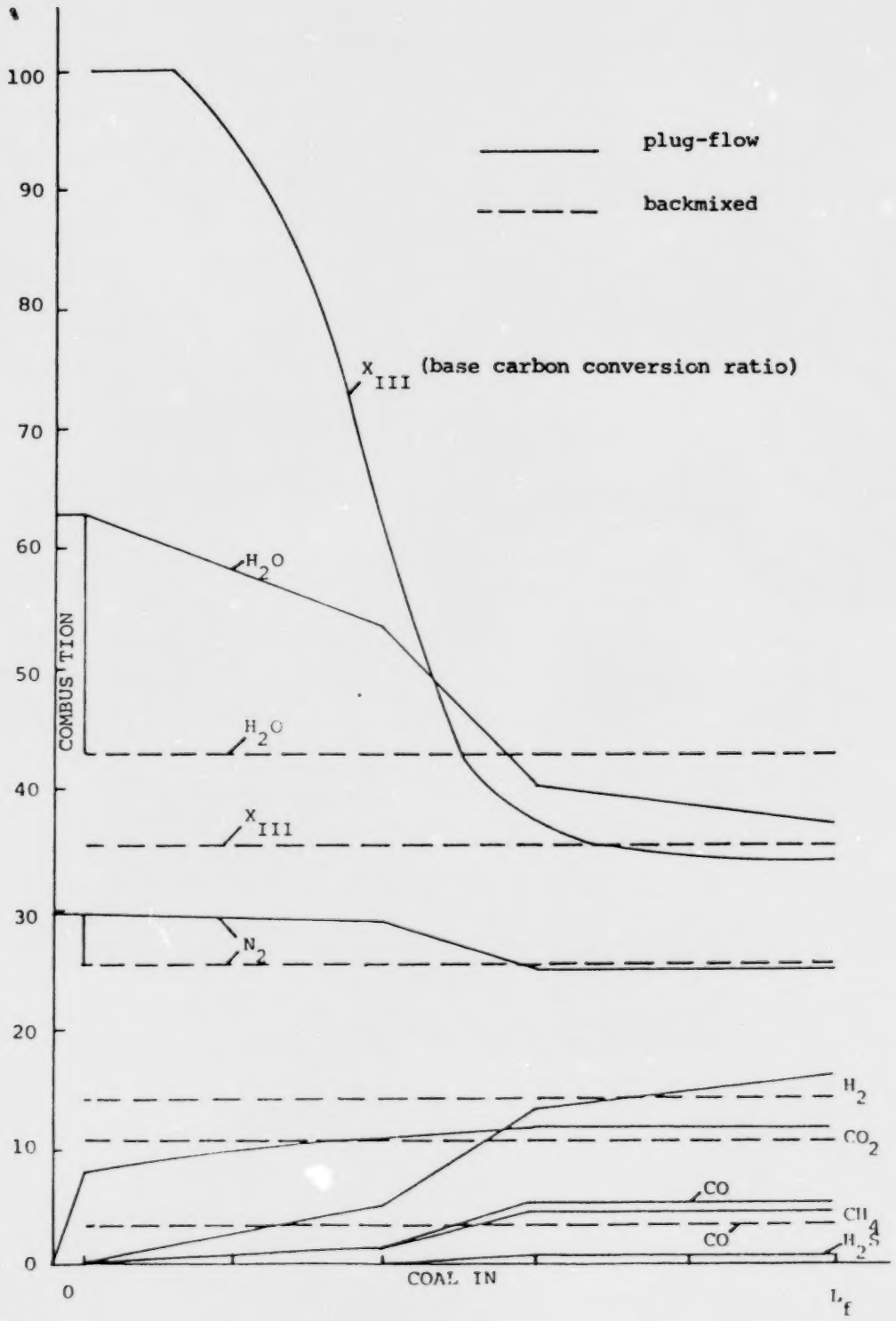


Figure 4 Comparison of predictions made by the plug-flow and backmixed reactor models

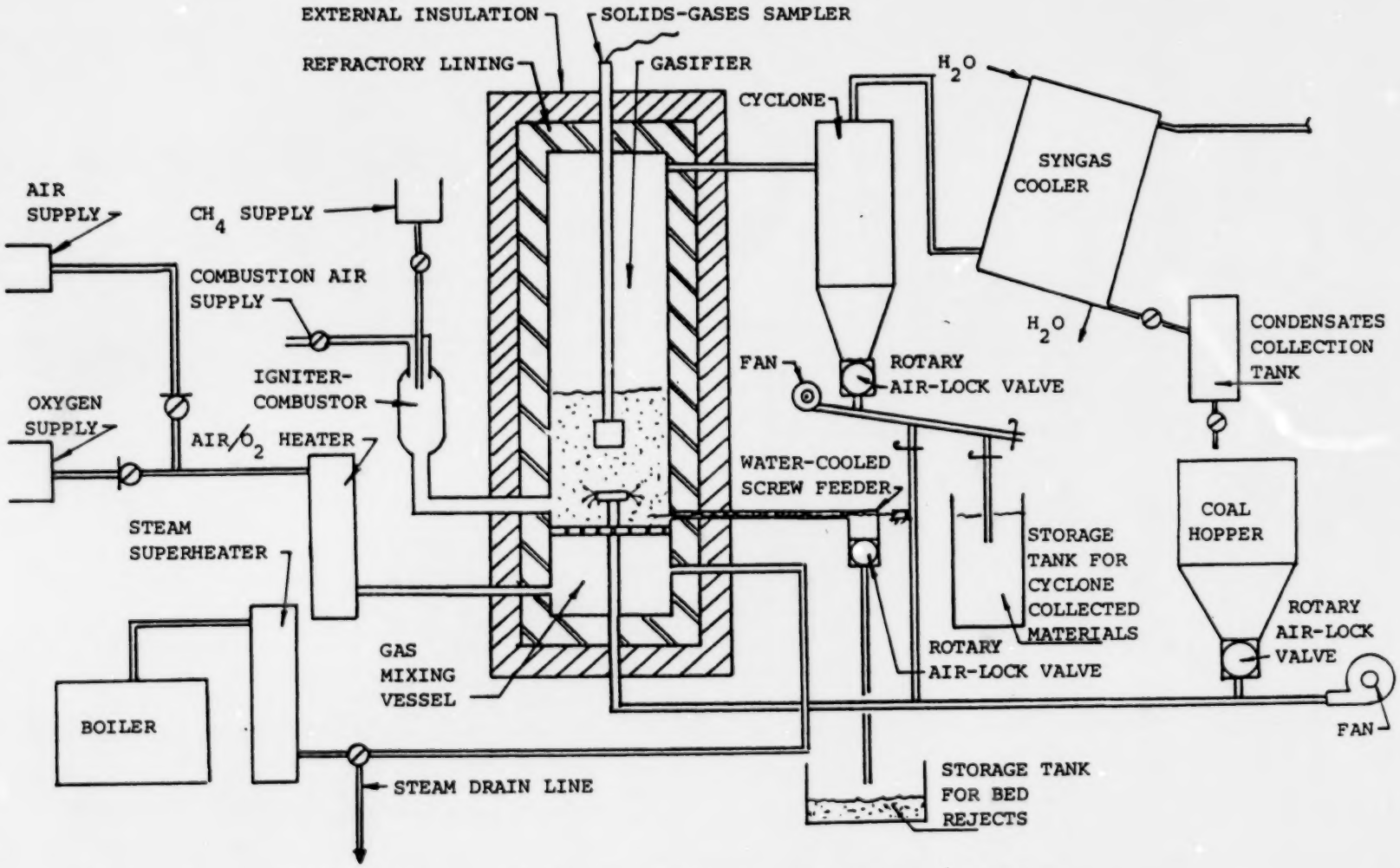


Figure 5 Schematic diagram of possible layout of suggested design

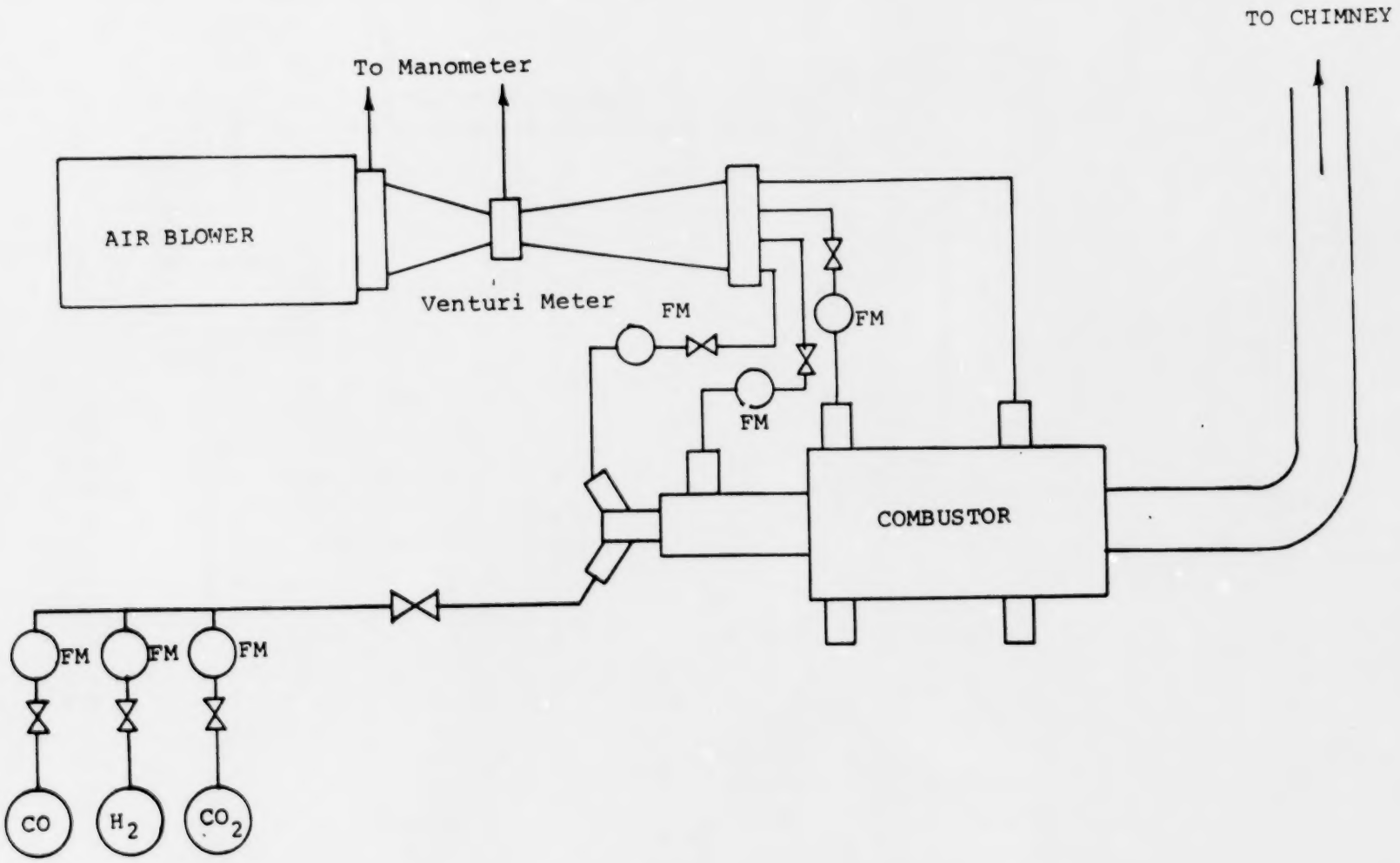


Figure 6 Schematic Diagram of Combustor Test Rig Layout

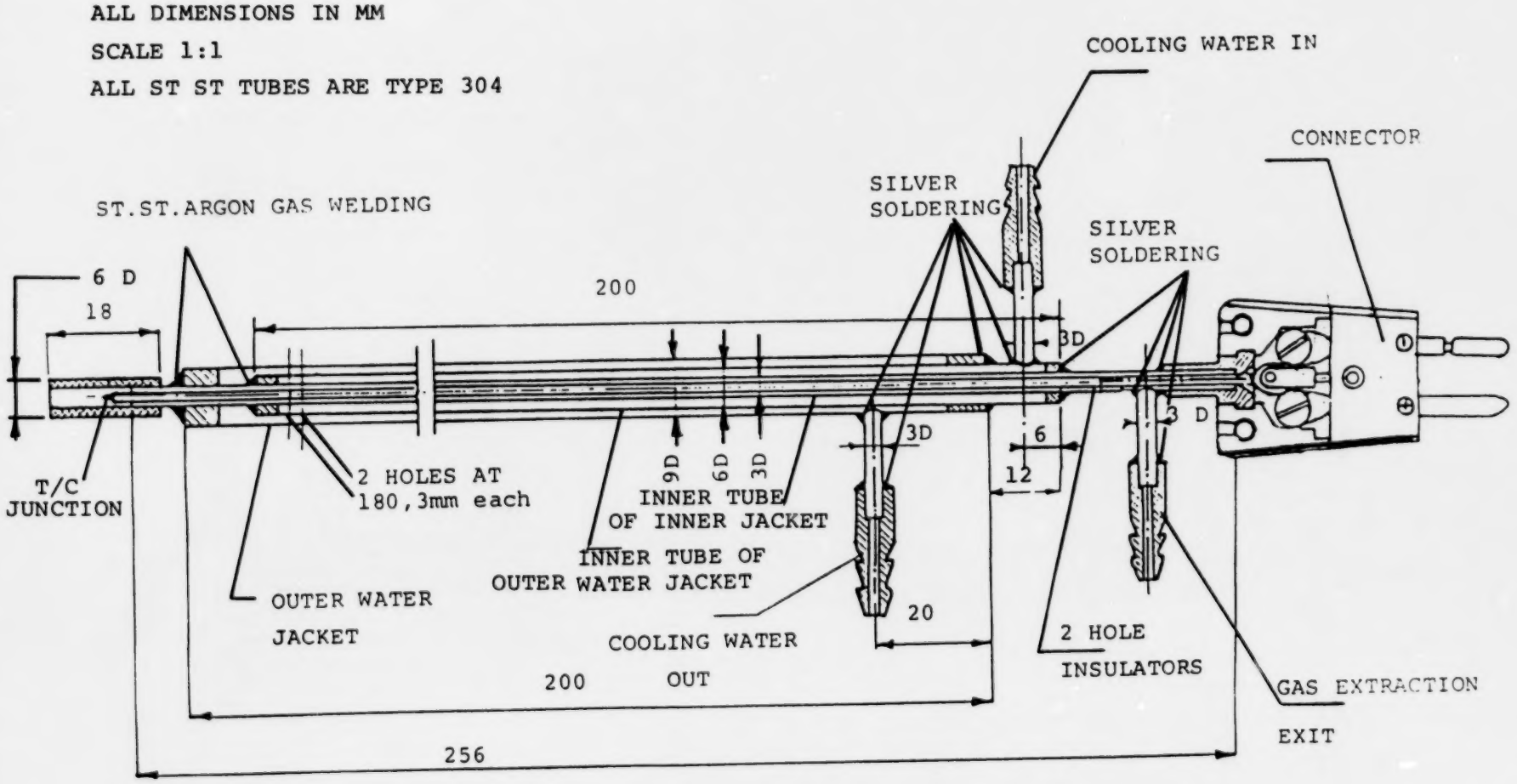


Figure 7 Suction pyrometer probe

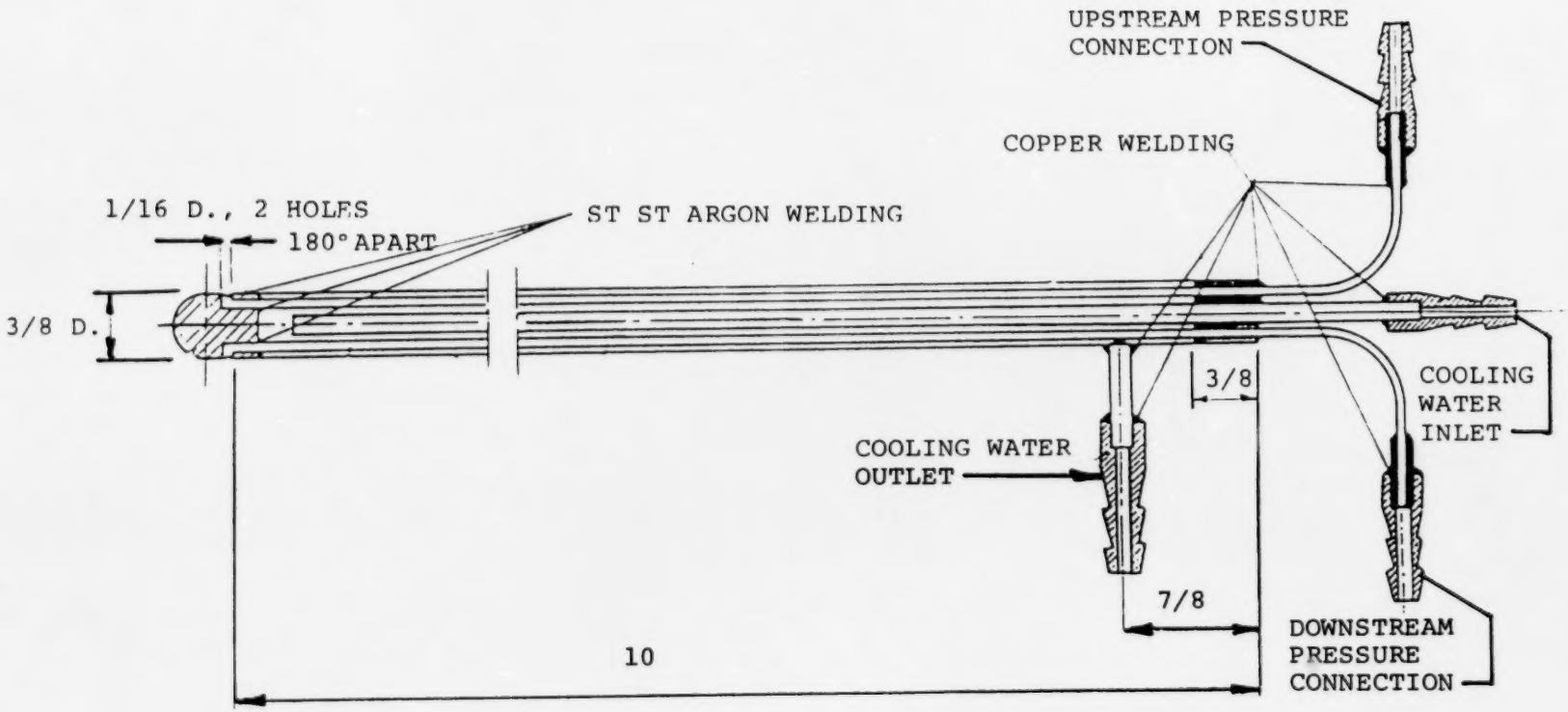


Figure 8 Two-hole probe

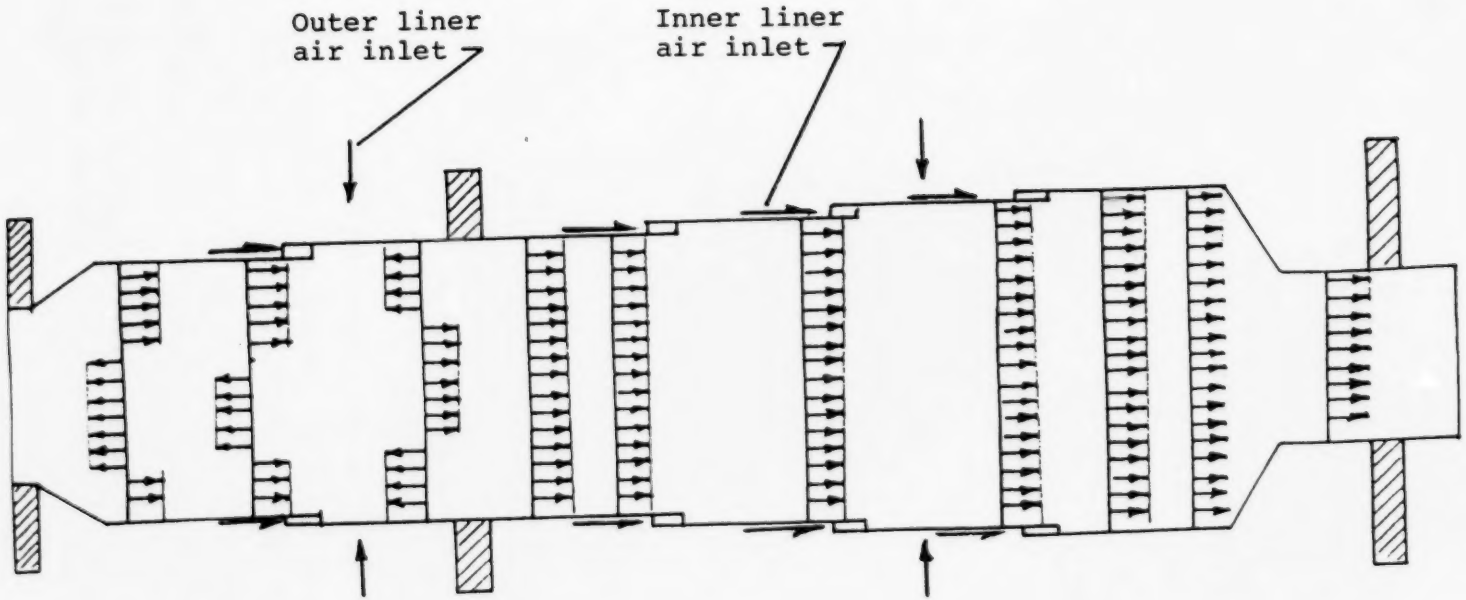


Figure 9 Typical results of flow direction measurements using the two-hole probe

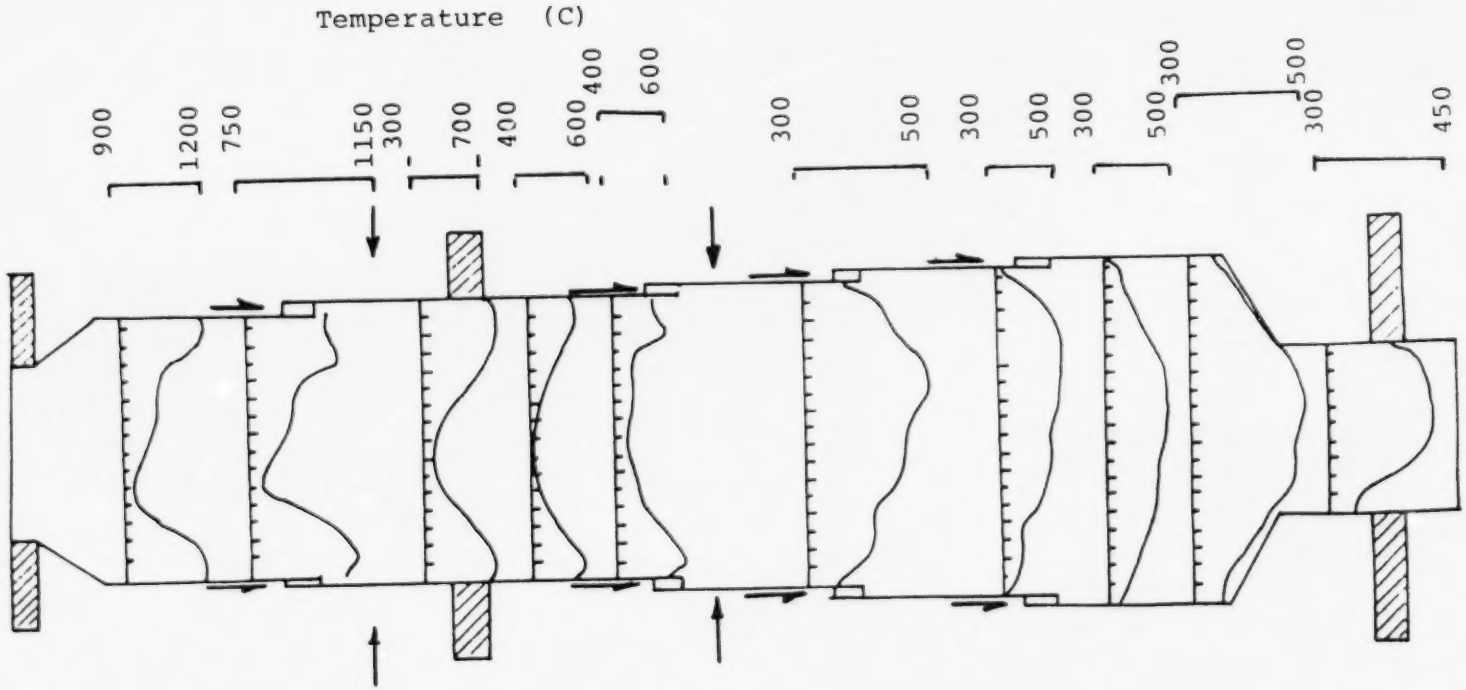
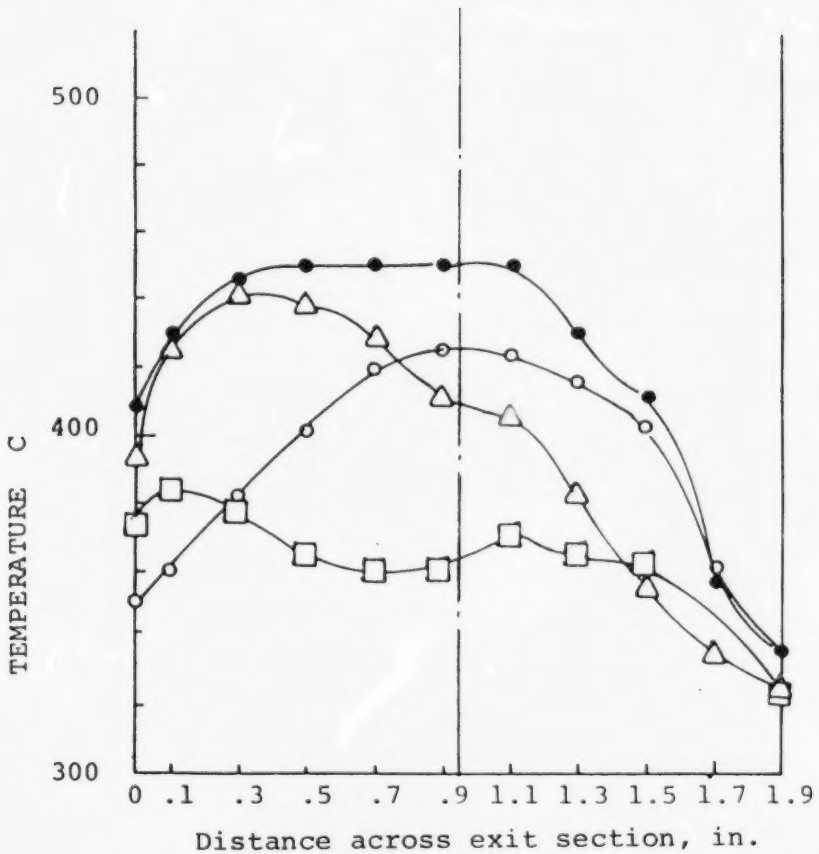


Figure 10 Temperature distribution across combustor at ten sections

A/F=12.98 by volume; -21.45 by mass

Air flow rate= 103 kg/hr

Heating value of fuel= 10.4 MJ/m^3 (283 Btu/ft^3)



—□— A/F = 17.1 by vol., -28.2 by mass; m_a = 103 kg/hr

—△— A/F = 14.7 by vol., -24.2 by mass; m_a = 88.5 kg/hr

—○— Same overall conditions as —△— but with different distribution of dilution air inlet

—●— A/F = 12.9 by vol., -21.5 by mass; m_a = 103 kg/hr

Figure 11 Radial temperature distribution at combustor exit section for four different conditions

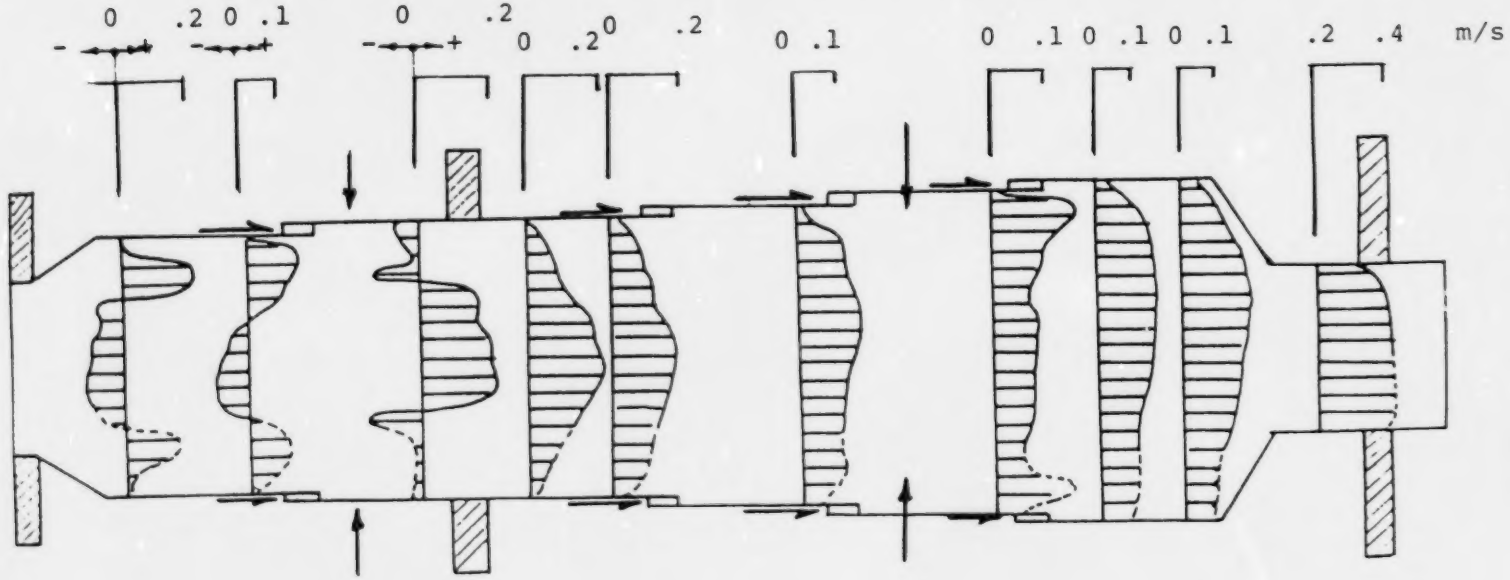
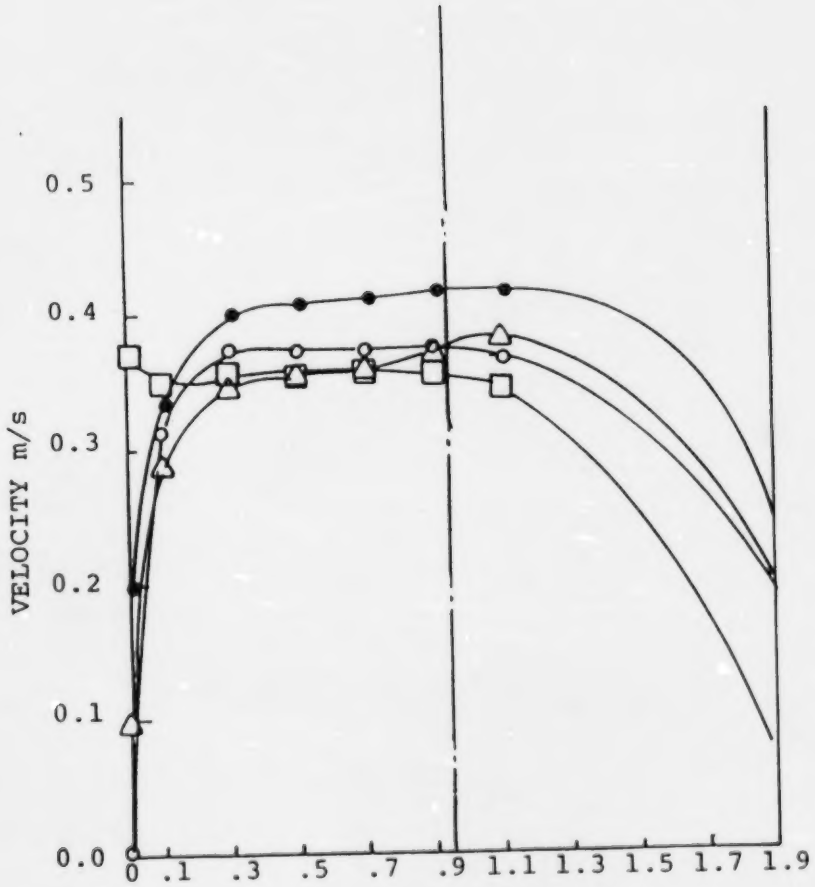


Figure 12 Axial velocity component distribution across combustor at ten sections

A/F = 12.9 by volume; -21.45 by mass

Air flow rate = 103 kg/hr

Heating Value = 10.4 MJ/m^3 (283 Btu/ft^3)



—□— A/F = 17.1 by vol., -28.2 by mass; m_a= 103 kg/hr

—△— A/F = 14.7 by vol., -24.2 by mass; m_a= 88.5 kg/hr

—○— Same overall conditions as —△— but with different distribution of dilution air inlet

—●— A/F = 12.9 by vol., -21.5 by mass; m_a= 103 kg/hr

Figure 13 Axial velocity component distribution at combustor exit section for four conditions

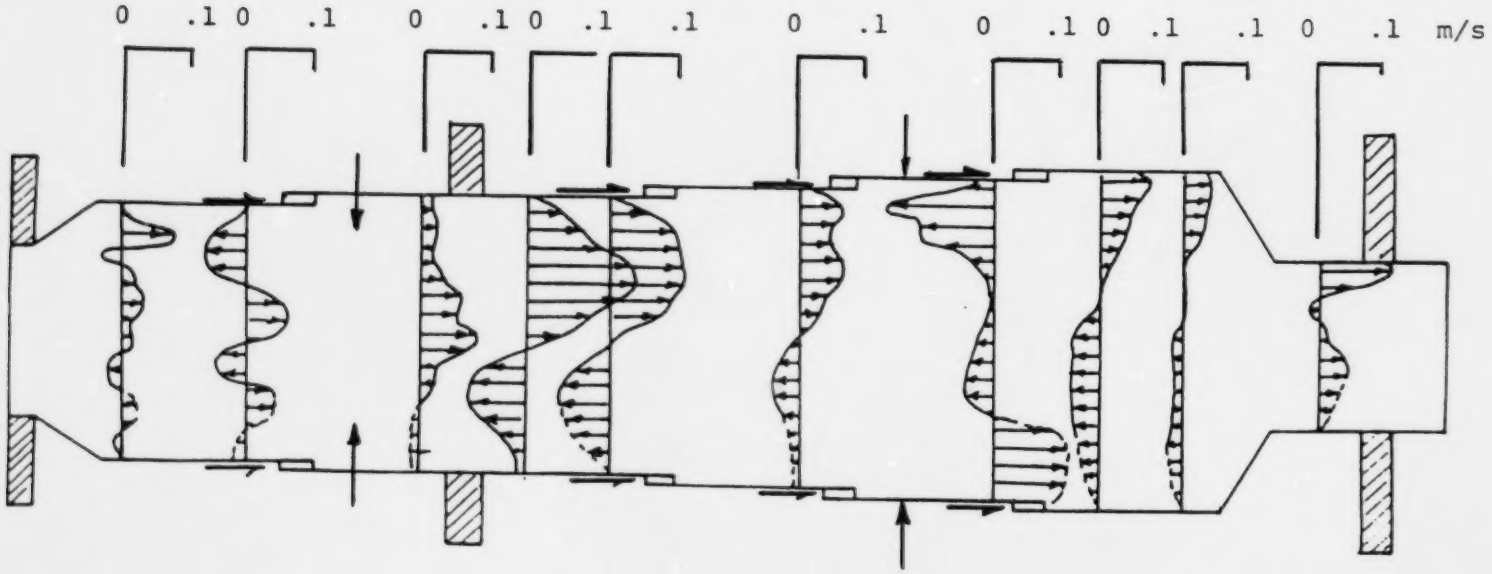
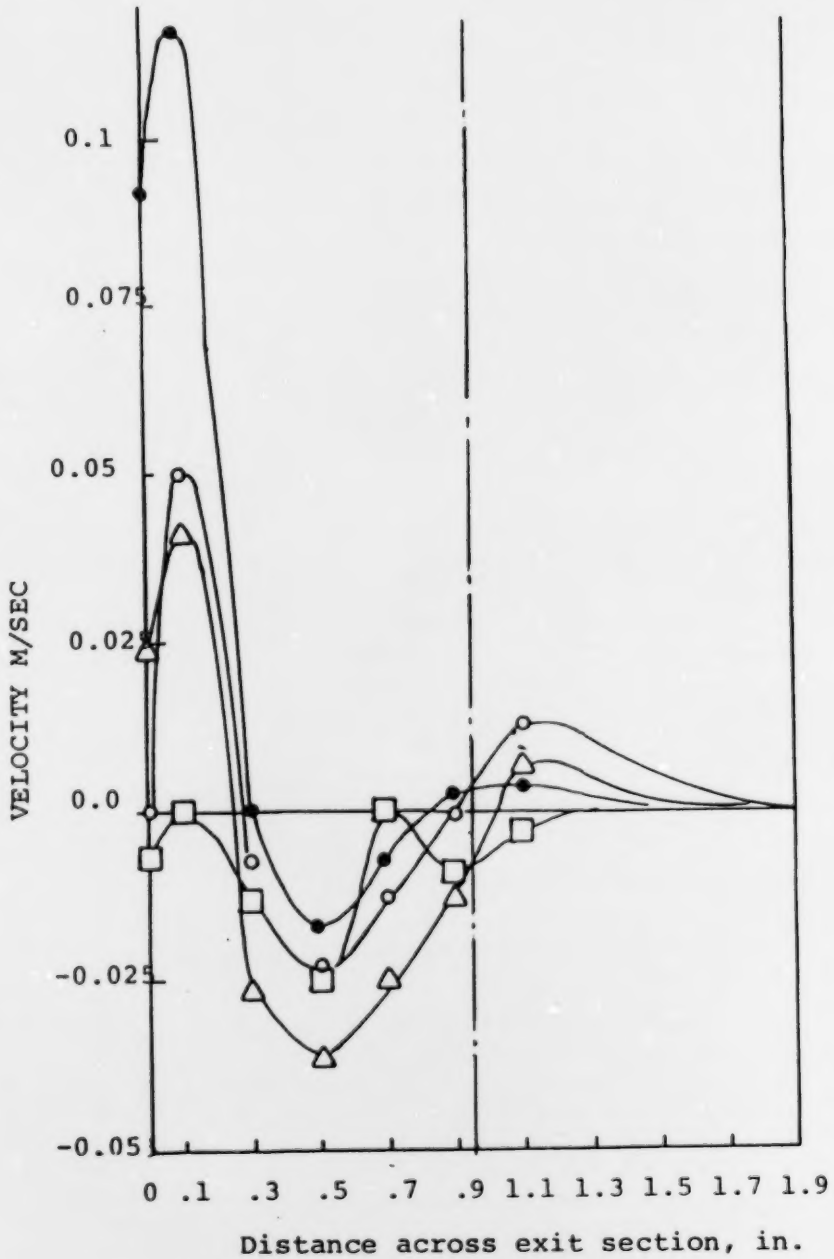


Figure 14 Radial velocity component distribution across combustor at ten sections

A/F = 12.98 by volume; -21.45 by mass

Air flow rate = 103 kg/hr

Heating value = 10.4 MJ/m^3 (283 Btu/ft^3)



—□— A/F = 17.1 by vol., -28.2 by mass; m_a = 103 kg/hr

—△— A/F = 14.7 by vol., -24.2 by mass; m_a = 88.5 kg/hr

—○— Same overall conditions as —△— but with different distribution of dilution air inlet

—●— A/F = 12.9 by vol., -21.5 by mass; m_a = 103 kg/hr

Figure 15 Radial velocity component distribution across combustor at exit section

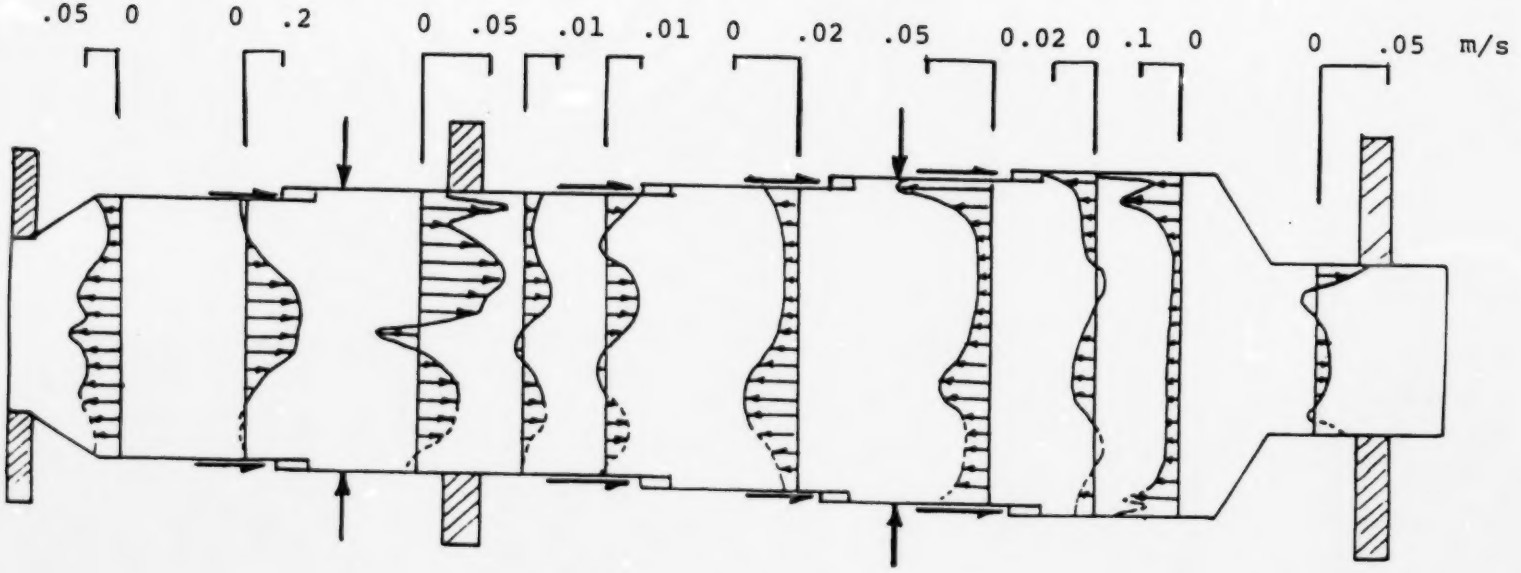
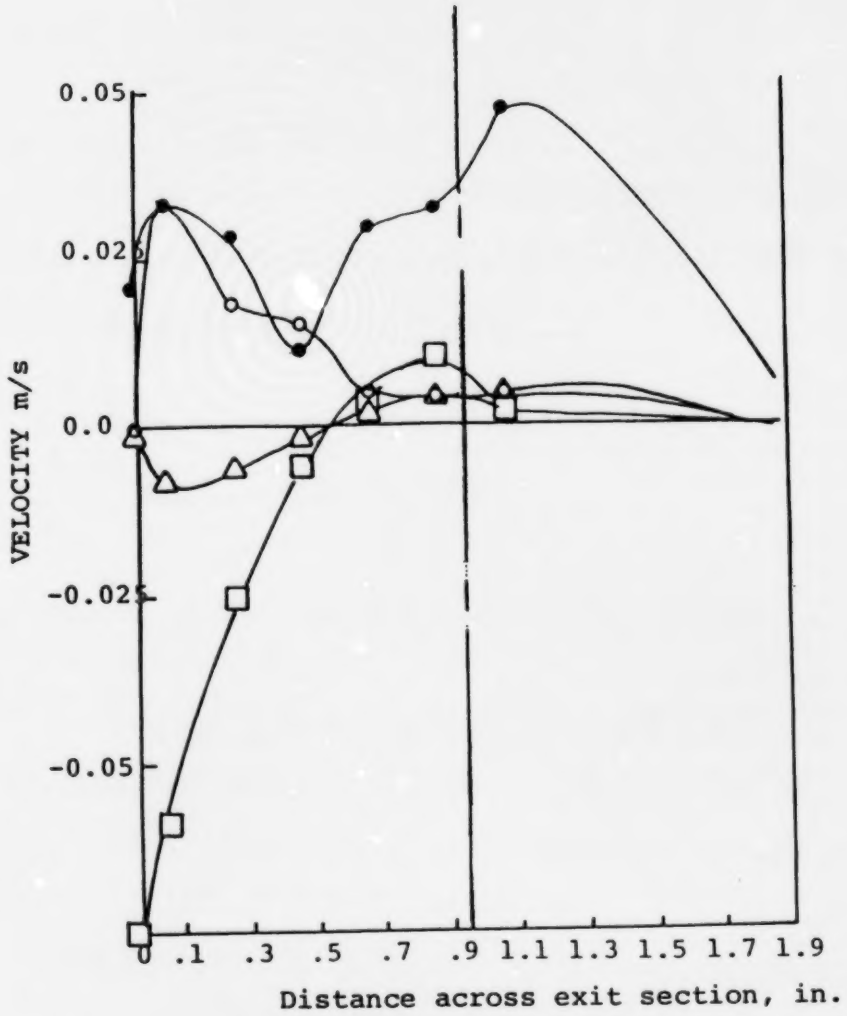


Figure 16 Tangential velocity component distribution across combustor at ten sections

A/F = 17.1 by volume; -28.2 by mass
Air flow rate = 103 kg/hr
Heating value = 10.4 MJ/m^3 (283 Btu/ft^3)



—□— A/F = 17.1 by vol., -28.2 by mass; m_a= 103 kg/hr

—△— A/F = 14.7 by vol., -24.2 by mass; m_a= 88.5 kg/hr

—○— Same overall conditions as —△— but with different distribution of dilution air inlet

—●— A/F = 12.9 by vol., -21.5 by mass; m_a= 103 kg/hr

Figure 17 Tangential velocity component distribution across combustor at exit section

APPENDIX I

COMPARISON OF THE ANALYTICAL RESULTS DEVELOPED IN THIS RESEARCH AND EXPERIMENTAL RESULTS AVAILABLE IN THE LITERATURE

The various submodels that were used in the development of the algorithms were carefully selected; attention was given to ensuring that these models were supported by extensive experimental work. Due to this fact, the results obtained from the algorithms are expected to correlate experimental results well. To demonstrate this, one should develop an extensive and systematic experimental study specifically designed to meet this objective. With the difficulties encountered in the experimental work undertaken (see Chapter 3) this aim was abandoned; a comparison of the analytical results with the experimental results available in the literature was attempted instead.

A complete comparison was found to be impossible due to lack of reported experimental data. Experimental results or operational information from commercial units are very limited and, what is more important, whenever they are available they are not complete; missing basic information does not permit the selection of the necessary data (e.g. coal properties, independent process variables, etc.) for running the computer programs and generating the corresponding analytical results.

No information is available on certain aspects of FBG performance, which then cannot be checked, e.g. turndown ratio of the FBG and

optimization of its performance. To our best knowledge, no information has been published on results of a systematic performance optimization of any unit of any scale.

Tables I.1 and I.2 summarize data reported from operation of commercial FBG's or experimental studies on pilot plants compared with the range of corresponding optimum values obtained from the analytical algorithms. It is striking that very basic data are missing from these references, e.g. geometry of gasifier, coal properties, temperature of gasifying media, bed size etc. Under these circumstances, the best that can be done is to check the basic trends and the usual ranges of the important parameters (bed temperature, syngas composition, gasifying media feed rates, efficiencies etc.).

The ranges of the following parameters have been found close:

- oxygen feed rate, expressed in kmole/kg-coal (around 0.02),
- steam feed rate, " " " " ,
- fluidized bed temperature,
- composition of dry syngas (H_2O in syngas varies widely since it depends on the steam feed rate, which is also quite variable).

In general, smaller coal residence times and, correspondingly, lower carbon conversion ratios are reported; this signifies non-optimal operation of the unit. It may also suggest that to obtain the theoretical optimum output requires far too long residence times to be practical. As pointed out by Psofogiannakis,[I-1] the residence time can be reduced by up to 80% (by reducing the bed weight) with only minimal effect on the performance. Further examination of performance

dependence on residence times is recommended. In many cases, especially in commercial gasifiers, larger coal particles (of the order of a few mm) are gasified; the quality of fluidization of these large particles, not easily fluidizable according to Geldart's classification, is not discussed anywhere. Finally, very little is reported about particles' elutriation from the FBG's that can be of any use for comparisons.

References

I-1 Psofogiannakis, M. G. "Analysis and Optimization of the Steady State Performance of a Fluidized Bed Coal Gasifier", Ph.D. Thesis, Dept. of Mechanical and Aeronautical Engineering, Carleton University, Aug. 1982.

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TABLE I.1

EXPERIMENTAL AND ANALYTICAL RESULTS; AIR AND STEAM GASIFICATION

#	GROUP	SYM-BOL	UNIT	ANALYTICAL RESULTS [§]	HIGH-T WINKLER [1-2]	WESTINGHOUSE PDU [1-3]	WESTINGHOUSE PDU [1-3]	WESTINGHOUSE PDU [1-3]
1	GEOMETRICAL - PHYSICAL PARAMETERS	D_g	m	6				
2		L_T	m	40				
3		N_D	-					
4		D_h	m					
5		l_h	m	0.05				
6		Δt_{INS}	m	0.10				
7		κ_{INS}	kW/mK	0.0001				
8		h_o	kW/m^2K	0.20				
9		U	"	$.995 \times 10^{-3}$				
10		A_t	m^2	28.27				
11		A	"	728.26				
12		η_{CN}	-	0.90				

				LIGNITE-BITUMINOUS	LIGNITE	PITTSBURGH COAL	INDIANA 7 COAL	PITTSBURGH COAL
13	COAL PROPERTIES	CV_{COAL}	kJ/kg	15310-26320		32300	27900	
14		C	"	39.3-64.7		47.0	46.0	
15		H	"	2.6-4.2		45	40	
16		O	"	2.7-12.5				
17		N	"	0.57-0.71				
18		S	"	0.27-6.8				
19		H_2O	"	35.5-2.6	12-18	2	6	
20		ASH	"	9.9-18.3		6	8	
21		ρ_1	kg/m^3	1200				
22		$\bar{d}_p$	m	500				
23		f_o	-	0.9-10				
24		f_R	-	1.5				
25	INDEPENDENT PROCESS VARIABLES	W	kg	150000-425000				
26		m_c	kg/sec	2.5-22.5	0.5555	0.107	0.114	0.0923
27		M'_{O_2}	$kmole/kg_c$	0.013-0.018		0.0283	0.0287	0.040
28		M'_{N_2}	"	0.042-0.068		0.1064	0.1079	0.150
29		M'_{H_2O}	"	0.015-0.081		0.0158	0.010	0.0117
30		P_m	kPa	up to 2000				
31		T_a	K	up to 850				
32		T_w	K	"				

33	PERFORMANCE PARAMETERS	SYNGAS COMPOSITION CO_2	%	9-12		12.6	21.2	11.87*
34		CO	"	9-15		14.2	10.3	20.06
35		H_2	"	15-20		10.3	6.0	5.05
36		CH_4	"	1-4		2.0	0.9	0.46
37		N_2	"	25-35		55.2	53.4	62.55
38		H_2O	"	20-33		5.7	8.2	-
39		O_2	"	0		0	0	0
40		H_2S	"	< 1		0	0	- 0
41		CV_H	kJ/m^3					
42		CV_{HO}	"	4250-4800				
43		HC	"	450-3500				
44		P_H	MW	98-1250				
45		P_{HCOLD}	MW					
46		η_c	-	up to 100%				
47		η_H	-	93-98				
48		η_{HCOLD}	-	67-72				
49		T	K	1175-1250	1193	1273	1238	1282
50		Δp_d	kPa	4-50				
51		Δp_b	"	40-160		1.68	1.575	
52		p_g	"	up to ~ 1900	506.6	1690	1690	1687

53	DEPENDENT PROCESS VARIABLES	u_o	m/sec	0.4-1.1		0.91	0.95	0.74
54		u_2	"	0.6-3.3				
55		u_{pf}	"	0.4-0.5				
56		u_T	"	1.6-3.5				
57		L_f	m	up to 21		8	7.5	
58		τ	hrs	15-85		1-2	1-2	
59		ε_{mf}	-	0.55				

§ For optimal operation of the gasifier

* Dry syngas

TABLE I.2

EXPERIMENTAL AND ANALYTICAL RESULTS; O_2 AND STEAM GASIFICATION

#	GROUP	SYM-BOL	UNIT	ANALYTICAL RESULTS §	1-5 PILOT PLANT	1-5 PILOT PLANT	1-4 WESTING-HOUSE PDU	PILOT PLANT	PILOT PLANT	1-6 WINKLER	1-7 WINKLER (LEUNA)	1-7 BASF
1	GEOMETRICAL - PHYSICAL PARAMETERS	D_g	m	6				0.152	0.152			
2		L_T	m	40								
3		N_D	-									
4		D_h	m									
5		l_h	m	0.05								
6		Δt_{INS}	m	0.10								
7		κ_{INS}	kW/mK	0.0001								
8		h_o	kW/m^2K	0.20								
9		U	"	$.995 \times 10^{-3}$								
10		A_t	m^2	28.27								
11		A	"	728.26								
12		η_{CH}	-	0.9								

				BITUMINOUS	IRSHA BORODINO	IRSHA BORODINO	PITTSBURGH COAL	BITUMINOUS CHAR	BITUMINOUS CHAR	BROWN COAL	BROWN COAL	-
13	COAL PROPERTIES	CV_{COAL}	kJ/kg	26320								
14		ρ_1	kg/m^3	1200								
15		C	%	64.7				80	88	52.9		
16		H	%	4.19						4.7		
17		O	%	2.69						21.6		
18		N	%	0.71						0.7		
19		S	%	0.68						0.4		
20		H_2O	%	2.61	6.6	10.6				8.6		
21		ASH	%	18.30								
22		$\bar{d}_p$	µm	500	630-1000	630-1000		400	250			
23		f_o	-	0.9228				1	30			
24		f_R	-	1.5								
25	INDEPENDENT PROCESS VARIABLES	W	kg	30,000 - 250,000								
26		m_c	kg/sec	8-50			.0877	.00699	.01143	1.433-3.046	8.1818	.433
27		M'_{O_2}	$kmole/kg_c$	0.02			.0225	.0180	.0187	.0163	.01495	
28		M'_{N_2}	"	0	0	0	0	0	0	0	0	0
29		M'_{H_2O}	"	0.02-0.05			.0377	.0377	.02711	.0086	.01576	.0664
30		P_m	kPa	150-1000								
31		T_a	K	up to 850								
32		T_w	K	"								

33	PERFORMANCE PARAMETERS	SYNGAS COMPOSITION	CO_2	%	7.5-12	44.6*	36.5*	17.17*	8.9		13.8**	15.7**	10.7**
34			CO	%	30-50	14.8	19.1	49.05	14.8		48.2	44.4	46.1
35			H_2	%	29-32	37.7	41.2	29.81	13.2		35.3	36.0	37.2
36			CH_4	%	2-5	2.2	2.5	3.16	0.7		1.8	1.6	1.8
37			N_2	%	<1	0.4	0.3	0.3			0.9	0.8	3.4
38			H_2O	%	8-22								
39			O_2	%	0	0.3	0.4	0			0	0	0
40			H_2S	%	<1			0.5					
41			CV_{II}	kJ/m^3									
42			CV_{IIO}	kJ/m^3	8500-10500	7737	8675						
43			IIC	"	550-3600								
44			P_{II}	MW	200-1300								
45			P_{IICOLD}	"	164-1166								
46			η_C	-	up to 100%				0.593	0.63	0.906	0.80	
47			η_{II}	-	90-93						0.749[F]		
48			η_{IICOLD}	-	77-79								

49	DEPENDENT PROCESS VARIABLES	T	K	1300-1370	1323	1303	1240	1280	1249		1073-1173	
50		Δp_d	kPa	4-12								
51		Δp_b	"	40-90								
52		p_g	"	up to 910	1000	1000	997	2481	2316		101.3	101.3
53		u_o	m/sec	0.5-0.8			0.70					
54		u_2	"	1.3-2.5								
55		u_{pf}	"	0.4-0.5								
56		u_T	"	2.2-3.6								
57		L_f	m	1-14				0.87	0.87			
58		τ	hrs	6.5-20								
59		c_{mf}	-	0.55								

§ For optimal operation of the gasifier

* Dry syngas

** Dry, H_2S-free syngas

F Defined as $(\dot{m}\ CV)_{SYNGAS}/(\dot{m}\ CV)_{COAL}$